

Am

RESERVE COPY  
PATENT SPECIFICATION

971,966

DRAWINGS ATTACHED.

Date of Application and filing Complete Specification :  
Sept. 27, 1962. No. 36652/62.

Application made in United States of America (No. 206,271)  
on June 29, 1962.

Complete Specification Published : Oct. 7, 1964.

© Crown Copyright 1964.

971,966



Index at Acceptance :—C5 E1.

International Classification :—C 10 g.

COMPLETE SPECIFICATION.

Continuous Process for Cracking Hydrocarbon Oil.

ERRATA

SPECIFICATION NO. 971,966

- Page 2, line 90, after "that" insert "a"
- Page 3, line 20, for "provids" read "provides"
- Page 3, line 88, for "fluidizing" read "fluidized"
- Page 3, line 123, for "fluidizing" read "fluidized"
- Page 4, lines 11 & 12, for "combinde" read "combined"
- Page 7, line 57, for "debutalized" read "debutanized"
- Page 7, line 118, after "stream" insert comma

THE PATENT OFFICE,  
29th October, 1964

the design of the unit, it was customary to estimate a gasoline yield and coke producing rate and then calculate to obtain a heat balance with a particular combined feed ratio (which may be defined as the ratio of the combined volume of fresh feed and recycle oil relative to the volume of fresh feed) and/or the combined feed temperature which would appear to provide a "steady-state" operation.

However, the error in such a design was the assumption that "coke-make" or coke yield, that is, the weight per cent of coke produced based upon fresh hydrocarbon feed, was a function of conversion rather than catalyst circulation rate and catalyst residence time in the reaction zone. Those units which are designed to operate at low conversion levels and hence low coke rates were frequently short of heat so that feed

D 29759/1(11)/R.109 200 10/64 PI

catalyst cooler or water spray system associated with the regenerator. In each case, it has been found that the units, as instrumentized and controlled, will thermally stabilize themselves, even though product yields which follow may be far from optimum. There is the ever present demand and necessity to obtain from cracking units a maximum throughput and optimum product yields, and with improved control means being incorporated as a part of the present improved method of operation to prevent damage to the unit due to afterburning (that is undesirable high temperature burning of carbon monoxide to carbon dioxide within the light phase zone or flue gas outlet of the regenerator), it has been found advantageous to vary operating control means to establish and maintain higher temperature levels

[P] ]

# PATENT SPECIFICATION

971,966

DRAWINGS ATTACHED.

971,966



*Date of Application and filing Complete Specification :  
Sept. 27, 1962. No. 36652/62.*

*Application made in United States of America (No. 206,271)  
on June 29, 1962.*

*Complete Specification Published : Oct. 7, 1964.*

© Crown Copyright 1964.

Index at Acceptance :—C5 E1.

International Classification :—C 10 g.

## COMPLETE SPECIFICATION.

### Continuous Process for Cracking Hydrocarbon Oil.

We, UNIVERSAL OIL PRODUCTS COMPANY, a Corporation organised under the laws of the State of Delaware, United States of America, of No. 30 Algonquin Road, Des Plaines, Illinois, United States of America, do hereby declare the invention, for which we pray that a patent may be granted to us, and the method by which it is to be performed, to be particularly described in and by the following statement:—

The present invention is directed to an improved method for operating a fluid catalytic cracking unit for the production of gasoline from heavier hydrocarbon oils and more specifically to a method for effecting improved product yields from a hydrocarbon charge stream.

It has been conventional for design and operating engineers to design and to operate fluid catalytic cracking units as "heat balanced" units. In other words, with prior data and experience as a basis for the design of the unit, it was customary to estimate a gasoline yield and coke producing rate and then calculate to obtain a heat balance with a particular combined feed ratio (which may be defined as the ratio of the combined volume of fresh feed and recycle oil relative to the volume of fresh feed) and/or the combined feed temperature which would appear to provide a "steady-state" operation.

However, the error in such a design was the assumption that "coke-make" or coke-yield, that is, the weight per cent of coke produced based upon fresh hydrocarbon feed, was a function of conversion rather than catalyst circulation rate and catalyst residence time in the reaction zone. Those units which are designed to operate at low conversion levels and hence low coke rates were frequently short of heat so that feed

heaters were provided to furnish the thermal requirements. High conversion units, on the other hand, had the problem of too much coke-make for heat balance requirements, so that recycle was provided as a cooling stream to transfer the excess heat into the main column where such heat could be used to regenerate steam. In other cases, catalyst coolers were provided to preclude what was thought to be excessive regenerator temperatures. It was customary to operate regenerators below 620° C. on the presumption that a high temperature would reduce catalyst activity and result in short catalyst life.

It may also be pointed out in connection with the operation of conventional fluidized cracking units, that various upsets in a so-called "steady-state" operation can occur, as for example, there may be a loss of the recycle stream, an upset in feed preheat to the reactor, or an upset involving a catalyst cooler or water spray system associated with the regenerator. In each case, it has been found that the units, as instrumentized and controlled, will thermally stabilize themselves, even though product yields which follow may be far from optimum. There is the ever present demand and necessity to obtain from cracking units a maximum throughput and optimum product yields, and with improved control means being incorporated as a part of the present improved method of operation to prevent damage to the unit due to afterburning (that is undesirable high temperature burning of carbon monoxide to carbon dioxide within the light phase zone or flue gas outlet of the regenerator), it has been found advantageous to vary operating control means to establish and maintain higher temperature levels

[P] ]

in the unit, regenerator temperature levels being in the 620° C. to 677° C. range, with the metallurgy of the regenerator system, including internals, governing the

5 upper temperature limit.

It is therefore a principal object of the present invention to accomplish an improved continuous fluid catalytic cracking operation through effecting an increased

10 temperature differential between the reactor and regenerator and a high temperature level in the regeneration zone.

A further object of the invention is to provide that one or more independent operating variables are utilized in an integrated manner in the fluidized unit to insure a coke deposition on the catalyst which will result in an increased temperature spread between the contact zones and a

15 resulting regenerating zone temperature level at above about 620° C.

As is well known in the petroleum industry and as will be more fully pointed out hereinafter in connection with the accompanying diagrammatic drawing, a conventional fluid catalytic cracking unit makes use of a reactor, stripper, spent catalyst slide valve, regenerator, standpipe, regenerated catalyst slide valve, and a riser

20 line leading back into the reactor, with the catalyst flowing through the various sections of the unit, in the order named. A large fractionator or "main column" is utilized to receive the overhead cracked

25 vaporous product stream from the reactor. Such fractionator provides gasoline and other desired side-cut product streams therefrom, as well as recycle oil streams which are charged back to the reaction

30 zone. The mixture of fresh hydrocarbon feed stream and recycle, referred to as "combined feed", is vaporized upon contacting heated regenerated catalyst particles at the base of the riser line and effects the lifting of catalyst in the reactor.

35 The reactor products along with a minute quantity of entrained catalyst are introduced into the main column where, of course, the products are fractionated into the various overhead and side-cut streams in accord with their volatility.

Reactor temperature controls the regenerator slide valve to provide, in turn, variations in catalyst flow from the regenerator to the reactor. The spent catalyst valve, regulating catalyst flow from the reactor to the regenerator, is operated by catalyst level control means in the reactor.

There are various independent control variables that will effect the conversion of a hydrocarbon gas oil stream and/or combined feed which is introduced into the reaction zone. As used herein, the term

40 "conversion" is defined as the volume per

cent of the feed which is converted to materials lighter than the light cycle oil as separated from the main column, but corrected to take into account the amount of gasoline in the charge stream. The actual weight ratio of catalyst to fresh feed flow rates is called the catalyst to oil ratio (C/O ratio) and is an important variable in controlling the severity of the cracking reaction. With a given hydrocarbon feed rate, an increased catalyst to oil ratio produces an increased flow rate of catalyst through the reaction zone. Although less coke will be formed on each unit of catalyst in view of a shorter contact time in the reaction zone, the total coke rate increases. On the other hand, an increased or decreased contact time between the oil and catalyst can be accomplished by a change in the space velocity ("weight hourly space velocity" being defined as the weight of fresh feed charged per hour divided by the weight of catalyst within the reaction zone). Thus, a decrease in weight hourly space velocity means that lesser quantity of oil is contacting a given quantity of catalyst per hour, or for a given time, such that there is an increase in conversion by reason of a longer time of contact with the catalyst. Conversely, an increased space velocity provides for the passage of a greater quantity of oil through the reaction zone per unit of time such that, with other variables remaining constant, there may be less conversion and less carbon or coke deposition on the catalyst particles (lower coke/catalyst ratio).

70

75

80

85

90

95

100

Still another independent variable which will affect the reaction zone and the conversion of the feed stream is the amount of preheat provided for the fresh feed stream to the reaction zone. The feed preheat may be varied directly by the amount of heat transferred to the feed stream by an oil heater or heat exchanger, or alternatively, by the quantity and temperature of the recycle stream which is combined with the fresh feed stream to provide the combined feed ratio.

Important independent operating variables may thus be considered the combined feed ratio, the combined feed temperature, and space velocity, although, of course, other variables such as the reactor pressure and the catalyst activity or quality will have an effect upon conversion. The catalyst circulation rate in the system may be readily varied; however, it is regulated responsive to a temperature controller in turn connecting with the reaction zone. Consequently, the circulation rate and the catalyst oil ratio are dependent variables in the operation of the commercial fluidized units.

115

120

125

The present invention thus is concerned

130

with a continuous process for cracking a hydrocarbon charge stream comprising fresh oil feed and recycle oil in the presence of subdivided catalyst particles wherein the hydrocarbon charge stream effects a fluidized contacting of the particles in a confined reaction zone, conversion products are separated from the contacted particles, separated catalyst particles containing a coke deposit are regenerated by contact in fluidized state with an oxygen-containing gas stream in a separate confined regeneration zone, gaseous combustion products are separated from regenerated catalyst particles and regenerated catalyst particles with a reduced coke content are returned to the reaction zone for contact with hydrocarbon charge stream, and in such process the present invention provides for effecting improved optimum product yields from the hydrocarbon charge stream with a high temperature level in the regeneration zone by the method which comprises, varying the temperature of the reaction zone and the contact time of the hydrocarbon stream with the catalyst therein responsive to the refractory characteristics of said hydrocarbon charge stream and effecting a coke deposition on the catalyst particles providing a temperature above about 620° C. in the regeneration zone when oxidizing the coke on the catalyst particles in the presence of a controlled oxygen-containing stream introduced to such regeneration zone, the introduction of such oxygen-containing stream being regulated to maintain a predetermined temperature differential between the gas outlet section and the catalyst-contacting section of the regeneration zone to minimize excess oxygen in the regeneration zone and to preclude excessive and uncontrolled after-burning in the upper portion of the regenerating zone.

45 In a more specific embodiment, the improved method of effecting superior product distribution from a hydrocarbon charge stream comprises, increasing the preheat temperature of the fresh feed stream to in turn provide an increased fresh feed temperature while reducing catalyst circulation rate, and to effect an increased temperature spread between the reaction and regeneration zones, such that the temperature level in the regeneration zone is above about 620° C. and up to the metallurgical limits thereof as the regeneration zone oxidizes the coke on the catalyst particles in the presence of a controlled oxygen containing stream introduced into the regeneration zone.

In still another specific embodiment, the method of effecting improved hydrocarbon product yields when operating in accordance with the present invention at a high

temperature level in the regenerator comprises, for a given fresh feed charge rate and temperature level, varying the combined feed ratio of fresh feed to recycle oils to adjust the heavy oil content thereof and simultaneously adjusting the reactor temperature and a temperature differential between the reactor and regenerator and a coke deposition on the catalyst particles providing a resulting temperature above about 620° C. upon oxidizing the coke on such particles in the presence of a controlled oxygen containing stream being introduced into the regeneration zone.

It should be pointed out, however, when one considers operating variables to increase the temperature spread between the contact zones, and the regeneration temperature, or cracking severity, that it is not sufficient to just make a change which alters conversion, since the important aspect of the present improved method of operating a fluidizing system includes effecting the desired product distribution at the same conversion. For example, a high severity operation with a small amount of recycle being used, will actually provide a lesser quantity of gasoline, due to the cracking of the gasoline into light gases in the reaction zone. For example, one can hold conversion constant, decrease coke-make and improve product yields, or one may increase conversion and hold the coke-make constant and effect improved product yields. The term "improved product yields" as used herein, means greater economic yield from the charge stream by reason of increasing the yield of gasoline and lighter materials which have more marketability.

In order to more easily refer to the operation of a fluid catalytic cracking unit and the operating variables in connection therewith, reference is made to the accompanying diagrammatic drawing and the following description thereof.

Fresh hydrocarbon feed passes through line 1 and control valve 2 to a pump 3 that in turn discharges into line 4 connecting with a feed preheater 5. The latter provides a predetermined or controlled variable preheating to the feed stream and passes it to the lower end of riser line 6 where heated catalyst particles combine therewith from standpipe 7, having control valve 8, such that a resulting vapor-catalyst mixture rises in an ascending fluidizing column to the lower end of the reactor 9. In the reactor 9, further fluidized contacting between the vaporous feed stream and the catalyst particles will take place in a relatively dense fluidized bed 10 within the lower portion of the chamber, although a major portion of the necessary cracking and contact with catalyst

particles takes place in the riser line 6 such that a shallow bed or low dense phase level of catalyst is utilized in chamber 9. Catalyst in an oil slurry may be combined with the fresh feed stream in the riser line 6 from line 11 having control valve 12, while in addition, recycle oil may be combined therewith by way of line 13 having control valve 14. As hereinbefore noted, the combined feed ratio will vary in accordance with the amount of recycle oils combine with the fresh feed stream to be introduced into the reaction zone.

At the upper end of the reactor 9, the catalyst particles are separated from the vaporous cracked reaction products by centrifugal separating means 15 and then transferred overhead by line 16 to the lower end of the fractionator or main column 17. Separated catalyst particles from the light phase zone in the top of the reactor 9 are returned to the dense phase bed 10 by a suitable dip-leg 18 and resulting catalyst particles with a coke deposition and occluded hydrocarbons settle from the lower portion of reactor 9 into a stripping section 19 such that they may pass countercurrently to a stripping gas stream introduced through line 20 having valve 21. Steam, nitrogen, or other substantially inert gaseous stripping medium may be utilized in the stripping section to effect the removal of absorbed and occluded hydrocarbon vaporous components. Resulting stripped and coked catalyst particles move from the lower portion of stripping section 19 into the standpipe 22, having control valve 23, such that they may be transferred at a controlled rate to the regenerator 24.

In the regenerating chamber 24 the carbonized catalyst particles are subjected to oxidation and carbon removal in the presence of air being introduced by way of distribution grid 25. The oxidizing air stream enters the regenerator by way of line 26, valve 27 and blower 28, which in turn connects by way of line 29 to air heater 30. In the present embodiment, the blower is indicated as connecting directly with the lower end of the regenerator 24 and the pipe grid distributor 25. The air heater 30 is utilized only to heat the air during the initial startup procedure. A bypass line 31, having control valve 32, connects with the air line 29 in order to vent a portion of the air stream being introduced to the system from blower 28 and thus regulate the quantity of air actually being introduced into the lower end of the regenerator 24, as will be more fully described hereinafter.

In the lower portion of the regenerator 24, a fluidized dense phase bed 33 provides for hindered settling contact between the coked catalyst particles and the oxidizing air stream while, in the upper portion of the chamber, a light phase zone permits the separation of catalyst particles from the flue gas stream being discharged by way of line 34 and valve 35. Suitable centrifugal separating means 36 provides for removing entrained catalyst particles from the combustion product stream and returns them by way of dip-leg 37 to the lower dense phase bed 33. A suitable silencing means 38, connecting with line 35, serves to reduce the noise level of the combustion gas stream passing to the outlet stack 39.

At the main column 17, the cracked product vapor stream is fractionated to provide a desired overhead gasoline vapor stream passing by way of line 40 and valve 41 to suitable gas concentration equipment, not shown in the present drawing. In addition various side-cut and recycle streams may be taken from the side of the fractionator 17 in accordance with desired molecular weight products. A light recycle oil stream is indicated as being taken from column 17 by way of line 42 and side-cut accumulator 43. An overhead vapor return line 44 returns uncondensed vapors from accumulator 43 to the main column 17 while a condensed light oil fraction is taken from the lower end of the accumulator 43 by way of line 45, pump 46 and line 57 having valve 48. Similarly, a heavier recycle oil stream may be taken from the main column by way of line 49 and side-cut accumulator 50. An uncondensed vapor stream may be returned to the main column by way of line 51 while a condensed heavy oil fraction passes from the lower end of accumulator 50 by line 52, pump 53, and line 54 having valve 55. Suitable lines, not indicated in the present drawing, may be utilized to transfer the light and heavy recycle oil streams from lines 47 and 54 to the recycle inlet line 13 which in turn connects with catalyst riser line 6, such that recycle oil may be combined with fresh hydrocarbon feed stream from line 1 to provide a desired combined feed ratio to the reactor 9.

A bottom slurry oil, containing catalyst particles which were entrained with the vapor product stream in line 16, may be transferred by way of the lower outlet line 56, pump 57, and line 58 to a suitable catalyst settling chamber 59. The settler 59 serves to provide a substantially particle free clarified oil stream overhead by way of line 60 and valve 61, while at the same time effecting the return of the catalyst containing slurry stream through control valve 12 to transfer line 11 such that catalyst particles may be returned to the fluidized bed by way of riser line 6. A

line 62, with control valve 63, is utilized to pass a portion of the bottoms stream from the main column 17 through a heat exchanger 64 such that heated oil may be returned by way of line 65 to the lower end of the main column 17 and provide a heat supply to such column.

In order to maintain control of the fluidized unit, control instrumentation means are associated with the reaction and regeneration zones to maintain appropriate dense phase bed levels in such zones and a catalyst circulation rate between such zones. At the reactor 9, a level controller LC, with level indicating taps 66 and 67, is connected with the side wall of the reactor. A control line 68 from controller LC connects with the slide valve 23 in the contacted catalyst standpipe 22 and provides means for maintaining a desired dense phase bed 10 level and quantity of catalyst in the lower portion of the reactor 9. Generally, the slide valve control means, such as used in connection with slide valve 23, are pneumatic in operation such that the level control means LC may be of any conventional type suitable to regulate the pneumatic motor control means of the valve, although electrical control and motor means may be used.

Within the upper portion of reactor 9, a temperature indicating means 69 connects through control line 70 to a temperature controller TC and control line 71 which in turn connects with an air piston or other motor means providing for the regulation of the slide valve 8 in the regenerated catalyst standpipe 7. Thus, a variable quantity of hot regenerated catalyst may be withdrawn from standpipe 7 to pass into riser line 6 and enter the reactor chamber 9 in accordance with variations in temperature in the latter zone, as provided for by the temperature sensitive means 69 and controller TC. A pressure sensitive means 72 is also positioned in the upper portion of the reactor 9 while a separate pressure indicating means 73 is positioned in the upper portion of the regenerator 24. Such indicating means are connective, through the respective lines 74 and 75, to a differential pressure regulator DPR in order to provide means for maintaining a desired differential pressure between the two separate contacting zones. The differential pressure regulator DPR connects through control line 76 with the control valve 35 so as to regulate the flue gas flow through line 34 and in turn vary the internal pressure within the upper portion of the regenerator 24, whereby a predetermined pressure difference may be maintained between the reactor and the regenerator. Generally, the pressure differential between the two zones, of the

order of about 0.42 kg/cm<sup>2</sup>, is necessary to permit the maintenance of suitable pressure differentials across the slide valves in the two standpipe lines and a continuous circulation of catalyst particles between two separate zones. Pressures in the fluidized unit are, of course, relatively low, being generally less than about 2.1 kg/cm<sup>2</sup> above atmospheric pressure because of the large size of the vessels involved. Pressure variation in the reactor may be an operating control variable, but because of structural and mechanical aspects, it is preferable to utilize low super-atmospheric pressures up to about 1.76 kg/cm<sup>2</sup> which are merely sufficient to insure adequate differentials and proper flow between various portions of the unit.

An improved control means integrated in connection with high temperature operations and indicated in the present diagrammatic embodiment, is the use of a differential temperature control means between the dense phase and light phase zones of regenerator 24 so as to regulate the quantity of air being introduced into the regeneration zone. Temperature indicating means 77 and 78, within the upper and lower portions of the regenerator 24, connect through the respective transmission lines 79 and 80 to a differential temperature controller DTC, which in turn connects through control line 81 with the valve 32 in the air vent line 31. Thus, where the temperature differential between the upper and lower portion of the regenerator exceeds a predetermined difference, as set within the temperature controller DTC, then valve 32 is adjusted to bypass a greater portion of the air passing through line 29 and effect the reduction of air being introduced by way of distributing grid 25 into the lower portion of the regenerator. Persons familiar with the operation of fluidized cracking units realize that uncontrolled after-burning can be a major problem inasmuch as the high temperature resulting from the oxidation of carbon monoxide to carbon dioxide in a dilute phase and the cyclone separators at the outlet of the regenerator may cause severe mechanical damage. In the prior operation of most present day units, it has been customary to provide for spray water and/or heat exchangers as cooling means, or to utilize flue gas analysis and accompanying control apparatus to severely reduce the oxygen introduction into the regeneration zone and as means for controlling the regenerator temperature to below about 607° C. However, the use of spray water has proved to be undesirable because of (1) damage to the catalyst by sintering or breakage by thermal shock of the particles, and (2) its nullifying effect on feed

preheat. Prior attempts to reduce the flow of the air stream and the introduction of oxygen into the regeneration zone has required sensitive means for detecting small oxygen concentrations and adjustment of the air introduction rate, all of which has been difficult from the practicable aspects. In the present invention where it is desired to have cracking operations which maintain a high temperature level in the regenerator, in other words, above about 620° C., it is similarly desirable to have means for precluding the problems of after-burning. It has been found that a temperature differential between the dense and dilute phases or between the dense phase and the flue gas outlet line is a very sensitive indicating means to show variations in the oxygen content of the gas leaving the dense phase itself. With this procedure, after-burning may be controlled such that the regenerator temperatures can be maintained at a steady-state level. In practice, it has been found that from 1% to 2% of the air blower output may be vented and that a temperature rise, over and above the predetermined differential set in the DTC may be utilized to control valve 32 and the vent rate through line 31. At the present time, it is generally necessary to determine by trial and error on any particular unit that temperature differential or rise which, when maintained constant, will regenerate the catalyst to lower the coke level on the catalyst to a desired lower level, generally 0.3–0.4 weight per cent, and maintain the free oxygen concentration in the flue gas below 0.2%.

In order to evaluate the present improved operation, utilizing high temperature levels in the system, one may consider two different charge stocks: an "A" stock which is of paraffinic base having a high "K" value, also called characterization factor, and is low in metals and Conradson carbon; as well as "B" stock which has a low "K" value and is high in contaminants. In a fluid catalytic cracking unit being operated and controlled in a manner as hereinbefore set forth in reference to the accompanying drawing, with a combined feed ratio of 1.5, a weight hourly space velocity equal to 2.0, and fresh feed temperature at 205° C. to the reactor riser line, the reactor temperature being adjusted to provide a 65 volume per cent conversion to gasoline, then the resulting steady-state operation provides a reactor temperature of 477° C., regenerator temperature of 593° C. and 7.7 weight per cent of coke. With the stock "B" being charged to the unit under the same conditions and with reactor temperature adjusted to provide a 65 volume per cent conversion to

gasoline, there is found a reactor temperature of 491° C., a regenerator temperature of 649° C. and coke amounting to 8.3 weight per cent. The 0.6 weight per cent difference in coke-make is due to sensible heat in the reactor and/or regenerator vapors.

From the foregoing figures it may be noted that the "poorer" charge stock "B" can be cracked in the conversion unit at a high temperature conversion level to give a relatively high 65 per cent conversion and there will be no problems in maintaining a steady-state operation as long as the 649° C. temperature in the regenerator is below the limits of any metallurgical problems encountered in the regenerator section of the system. On the other hand, it becomes apparent that the charge stock "A" could be treated in the system to provide substantially greater yields of gasoline or light materials by effecting changes in the operating variables so as to provide an increased regenerator temperature of the order of 649° C. In other words, there may be an increase in the preheating of the feed stream, say to the order of 371° C. rather than the 204° C. feed temperature, an increase in regenerator temperature say of the order of 42° C. and provide a reduction in coke-make from about 7.7 to less than 5.5 weight per cent; conversion is maintained constant by increasing reactor temperature. The lesser amount of coke-make indicates an increased amount of gasoline and lighter products. The increased feed preheat and the increased regenerator temperature level permits the regenerated catalyst slide valve to be closed down and to lower the catalyst circulation rate in the system so that actually catalyst enters the reaction zone at a reduced rate and a higher temperature with the resulting effect that there will be a longer contact time of the catalyst particles but a reduction in total carbon deposition due to reduced catalyst circulation rate. The "trade" between coke-make and more valuable product yields effects a very desirable improvement in the economics of the unit. It is also important to note that a fluid catalytic cracking unit can be varied in its operation in accordance with the refractory characteristics of the feed and, in fact, evaluate a particular charge stock in terms of reactor and regenerator temperature levels such that with the better charge stocks it is possible to consistently improve yields by operating to maintain a high temperature level in the regenerator as long as the high temperature level precludes metallurgical damage to the unit.

In another instance, a series of performance tests were conducted in a fluid cata-

lytic cracking unit in a manner which illustrates the advantage of a high temperature level in the system and an increased temperature differential between the reactor and regenerator zones. In the particular performance tests, three different levels of feed preheat were used, with conversion being maintained constant by adjustments in the reactor temperature. Specifically, the raw oil introduction rate was maintained at 44,510 hectoliters per day, the combined feed ratio was held constant for each performance test at 2.30, and catalyst content in the reactor or weight hourly space velocity was held substantially constant. In addition, in order to maintain uniformity during the tests as consistent as possible, there was a constant crude mix maintained in the charge to the feed preparation unit so that the raw oil feed rate was consistent and levels in the receivers and reboilers were held constant. In a Test No. 1, the fresh feed was held to a temperature of 212° C. to provide a combined feed temperature of 260° C. The resulting reactor temperature for 69.7 volume per cent conversion was 466° C. and the regenerator temperature 620° C. The catalyst circulation rate leveled off at  $2.031 \times 10^6$  kg per hour and coke make was 9.4 weight per cent. Debutanized gasoline was measured at 49.6 volume per cent while other product yields were determined to be in accordance with the quantities set forth in the following Table I, under the column head Test No. 1.

In a Test No. 2, the fresh feed was preheated to a temperature of 298° C. such that the combined feed temperature was 297° C. and the resulting reactor temperature and regenerator temperature were respectively 472° C. and 631° C. The catalyst circulation rate in this instance was  $1.724 \times 10^6$  while the coke-make rate was 8.6 weight per cent. The debutanized gasoline production was 50.1 volume per cent while other yields were in accordance with the quantities shown in Table I, under the column headed Test No. 2.

In a Test No. 3, the fresh feed was preheated to a temperature of 370° C. to provide a combined feed temperature of 327° C. and resulting reactor and regenerator temperatures, respectively, of 476° C. and 641° C. The catalyst circulation rate was  $1.497 \times 10^6$  kg per hour and the coke-make was 7.9 weight per cent. The debutanized gasoline was 51.2 volume per cent while other product yields were in accordance with the quantities set forth in the column under Test No. 3 of Table I.

Referring specifically to the Test No. 3 and the data in Table I, there is noted a delta T ( $\Delta T$ ) of 165° C., which is the difference between the reactor temperature

and regenerator temperature, that is higher than occurs in either of the other test operations. It may also be noted that there is a significantly greater gasoline yield in Test No. 3, together with less hydrogen production in the C<sub>3</sub> and lighter gases, showing that there is lesser hydrogen and coke-make which results in the increase of gasoline yield from the system. Actually, in the highest temperature level operation for the regenerator, as shown in Test No. 3, there is an increased coke deposition on the individual catalyst particles as such, due to a longer contact time in the reactor, and the resulting higher temperature level in the regenerator due to the oxidation of the coke from the catalyst particles. However, by reason of a substantially reduced catalyst circulation rate, the overall rate of producing coke is reduced in the system and the coke-make, as indicated as weight per cent of the feed, is a lower value in Test No. 3, than in the other two tests.

By way of summary, it may be noted from Table I that the coke reduction in turn appears as gasoline product even though the reaction temperature has been raised.

The weight fraction of C<sub>4</sub> and lighter components remain substantially constant, but the C<sub>3</sub> and C<sub>4</sub> fractions decreased with a corresponding increase in the C<sub>2</sub> and lighter fractions.

The olefin content of C<sub>3</sub> and C<sub>4</sub> fractions increase, with the largest change being in the C<sub>4</sub> fraction.

The isobutane content of the C<sub>4</sub> paraffins remains substantially unchanged as well as the properties of the liquid product.

The foregoing tests indicated changes in the regenerator temperature levels and product yields by varying the feed preheat, however, other independent control variables may be modified to raise regenerator temperature to just below the design maximum. For example, there may be a decrease in the yield of the main column bottoms stream (indicated in the drawing as the clarified oil stream passing by way of line 60), by recycling to the reactor a higher quantity of oil in the slurry bottoms stream (such as the stream passing by way of line 11 to riser 6 in the drawing).

In other words, any combination of control variables may be employed up to the regenerator temperature limit set by the metallurgical aspects of the regenerator portion of the unit, such that temperature may range up to the order of 677° C. High temperature, with good operating practices, provides no problem with catalyst sintering or deactivation and, in fact, regenerator operation appears to improve with respect to more uniform coke removal. Un-



controlled after-burning may be precluded at the high regeneration temperature levels by the hereinbefore described system of utilizing a temperature differential control means across the upper and lower portions of the regenerator chamber to effect the control of the quantity of air being introduced into the dense phase zone of the regenerator. Temperature differential becomes a sensitive measure of the quantity of oxygen present and by having insufficient oxygen present to permit the oxidation of carbon monoxide to carbon dioxide in the light phase zone of the unit, there is proper control to prevent a temperature runaway. Generally, a temperature differential of 14° C. as a maximum will permit a differential temperature controller to regulate the proportion of air to the regenerator by adjustment of the valve in the air vent line from the air blower, to in turn provide for a steady-state regenera-

tor operation which will preclude after-burning and at the same time effect a suitable reduction in the coke level to less than 0.5 per cent coke by weight and preferably to 0.3 to 0.4 per cent residual on the catalyst particles. Of equal importance, however, the oxygen content of flue gas remains at 0.0—0.2 per cent which reduces the oxidation rate of the cyclone separators.

The coke production rate and the catalyst circulation rate vary inversely with the regenerator temperature, as, for example, a "cool" regenerator operating temperature at say 593° C. results in a high coke rate for a given combined feed ratio and combined feed temperature. Stated another way, the coke rate opposes the direction in which the regenerator temperature moves, so that operations at high temperature levels in the regenerator are in the direction of improved product yields.

45

TABLE I

		Test No. 1	Test No. 2	Test No. 3
<i>Operating Conditions</i>				
	Fresh Feed Temp., °C. ...	212	298	370
	Combined Feed Temp., °C. ...	260	297	327
50	Combined Feed Ratio ...	2.30	2.30	2.30
	Reactor Temperature, °C. ...	466	472	476
	Regenerator Temp., °C. ...	620	631	641
	Δ T °C. (Temp. Differential between Reactor and Regenerator) ...	154	159	165
55	Δ TD °C. (Temp. Differential between Dense and Dilute Phases of Regenerator) ...	9	7	10
	Oxygen Content of Flue Gas mole % ...	0.1	0.1	0.1
	Reactor Pressure*, kg/cm <sup>2</sup> ...	1.055	1.055	1.055
60	Apparent Conversion, Vol.-% ...	69.7	69.7	69.7
	Corrected Conversion **, Vol.-% ...	69.2	68.9	69.6
	Catalyst Circulation Rate, kg/hour × 10 <sup>-3</sup> ...	2.031	1.724	1.497
<i>Product Yields</i>				
	C <sub>3</sub> and Lighter, Weight-% ...	7.6	7.8	9.1
65	Total C <sub>4</sub> Fraction, Weight-% ...	8.8	8.6	7.7
	Debutanized Gasoline**, Vol.-% ...	49.6	50.1	51.2
	Light Cycle Oil, Vol.-% ...	26.1	27.6	26.7
	Clarified Slurry, Vol.-% ...	4.2	3.5	3.7
	Coke, Weight-% ...	9.4	8.6	7.9
70	C <sub>3</sub> and Lighter, Mole-%			
	H <sub>2</sub> ...	17	16	14
	C <sub>1</sub> ...	26	27	29
	C <sub>2</sub> ...	19	22	24
	C <sub>3</sub> = ...	25	24	23
75	C <sub>3</sub> ...	13	11	10
		100	100	100
<i>Total C<sub>4</sub>, Liquid Vol.-%</i>				
	C <sub>4</sub> H <sub>10</sub> ...	46	50	53
	i-C <sub>4</sub> H <sub>10</sub> ...	43	40	37
80	n-C <sub>4</sub> H <sub>10</sub> ...	11	10	10
		100	100	100

\*Above atmospheric pressure.

\*\*Gasoline with 90% boiling at 196° C.

## WHAT WE CLAIM IS:—

1. A continuous process for cracking a hydrocarbon charge stream comprising fresh oil feed and recycle oil in the presence of subdivided catalyst particles, wherein the hydrocarbon stream effects a fluidized contacting of the particles in a confined reaction zone, conversion products are separated from the contacted particles, separated catalyst particles containing a coke deposit are regenerated by contact in fluidized state with an oxygen-containing gas stream in a separate confined regeneration zone, gaseous combustion products are separated from regenerated catalyst particles, and regenerated catalyst particles with a reduced coke content are returned to the reaction zone for contact with hydrocarbon charge stream, characterized by the method of effecting improved product yields from the hydrocarbon charge stream with a high temperature level in the regeneration zone, which comprises, varying the temperature of the reaction zone and the contact time of the hydrocarbon stream with the catalyst therein responsive to the refractory characteristics of said hydrocarbon charge stream and effecting a coke deposition on the catalyst particles which provides a temperature above about 620° C. in the regeneration zone when oxidizing the coke on the catalyst particles in the presence of a controlled oxygen-containing stream introduced to said regeneration zone, the introduction of said oxygen-containing stream being regulated to maintain a predetermined temperature differential between the gas outlet section and the catalyst contacting section of the regeneration zone to minimize excess oxygen in the regeneration zone and to preclude afterburning in the upper portion of the regenerating zone.
2. Process according to Claim 1, characterized by the method of operation which comprises preheating the fresh feed stream and varying the catalyst circulation rate to effect the desired coke deposition level on the catalyst particles.
3. Process according to Claim 1 or 2, characterized by the method of operation which comprises, varying the combined feed ratio of fresh oil feed to recycle oil to adjust the heavy oil content thereof and simultaneously adjusting reaction zone temperature to effect a coke deposition level on the catalyst particles.
4. Process according to one of the Claims 1 to 3, characterized in that the temperature and time of contact of the hydrocarbon stream with the fluidized catalyst in the reaction zone are adjusted so that the oxidizing of the resultant coke deposition on the catalyst particles in the regeneration zone effects a temperature between 620° and 677° C. in the contacting section of said regeneration zone, and the temperature differential between the gas outlet section and the contacting section of said regeneration zone is maintained at maximally 14° C.
5. Process according to Claim 4, characterized in that the temperature differential between the gas outlet section and the contacting section of the regeneration zone is maintained at a level which corresponds to a free oxygen concentration below 0.2% in the flue gas issuing from said outlet section.

J. Y. & G. W. JOHNSON,  
Furnival House,  
14—18 High Holborn,  
London, W.C.1,  
Chartered Patent Agents,  
Agents for the Applicants.





